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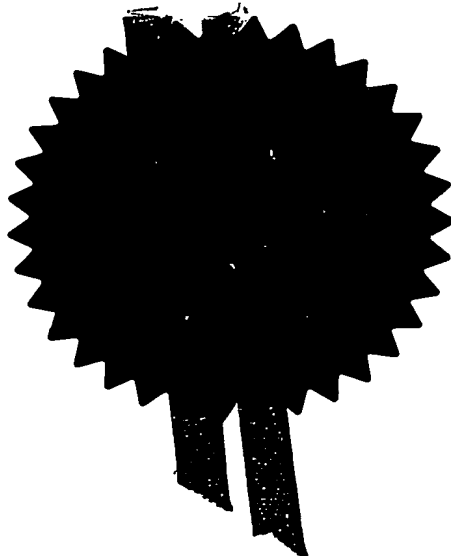
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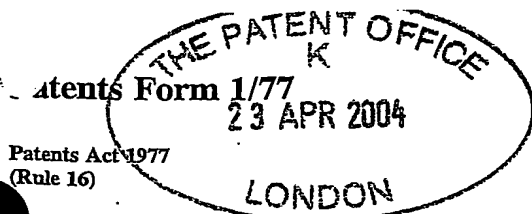


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*Stephen Hordley*

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## Request for grant of a patent

(See the notes on the back of this form. You can also get an explanatory leaflet from the Patent Office to help you fill in this form)

1. Your reference	31.13.84301		
2. Patent application number (The Patent Office will fill in this part)	0409103.9		23 APR 2004
3. Full name, address and postcode of the or of each applicant (underline all surnames)	1) Statoil ASA I&K IR PAT N-4035 8392722002 Stavanger Norway  2) Linde Aktiengesellschaft Abraham-Lincoln-Str. 21 D-65189 Wiesbaden Germany 571463009		
Patents ADP number (if you know it)			
If the applicant is a corporate body, give country/state of its incorporation	1) Norway 2) Germany		
4. Title of the invention	Natural Gas Liquefaction Process		
5. Name of your agent (if you have one)	Frank B. Dehn & Co.  179 Queen Victoria Street London EC4V 4EL		
"Address for service" in the United Kingdom to which all correspondence should be sent (including the postcode)			
Patents ADP number (if you know it)	166001 ✓		
6. Priority: Complete this section if you are declaring priority from one or more earlier patent applications, filed in the last 12 months	Country	Priority application number (if you know it)	Date of filing (day / month / year)
	DE	10344030.5	23/9/03
7. Divisionals, etc: Complete this section only if this application is a divisional application or resulted from an entitlement dispute (see note f)	Number of earlier UK application	Date of filing (day / month / year)	
8. Is a Patents Form 7/77 (Statement of inventorship and of right to grant of a patent) required in support of this request? Answer YES if: a) any applicant named in part 3 is not an inventor, or b) there is an inventor who is not named as an applicant, or c) any named applicant is a corporate body. Otherwise answer NO (See note d)	Yes		

# Patents Form 1/77

9. Accompanying documents: A patent application must include a description of the invention. Not counting duplicates, please enter the number of pages of each item accompanying this form:

Continuation sheets of this form

Description

19

Claim(s)

Abstract

Drawing(s)

5

10. If you are also filing any of the following, state how many against each item.

Priority documents

Translations of priority documents

Statement of inventorship and right to grant of a patent (Patents Form 7/77)

Request for a preliminary examination and search (Patents Form 9/77)

Request for a substantive examination (Patents Form 10/77)

Any other documents (please specify)

11.

I/We request the grant of a patent on the basis of this application.

Signature

Date 23 April 2004

12. Name, daytime telephone number and e-mail address, if any, of person to contact in the United Kingdom

Frank B Dehn & Co  
020 7206 0600

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84301.623

NATURAL GAS LIQUEFACTION PROCESS

5           The present invention relates to a method for the liquefaction of a hydrocarbon-rich flow.

10           Natural gas can be obtained from the earth to form a natural gas feed which must be processed before it can be used commercially. Normally the gas is first pre-treated to remove or reduce the content of impurities such as carbon dioxide, water, hydrogen sulphide, mercury, etc.

15           The gas is often liquefied before being transported to its point of use to provide liquefied natural gas (LNG). This enables the volume of gas to be reduced by about 600 fold, which greatly reduces the transportation costs. Since natural gas is a mixture of gases, it liquefies over a range of temperatures. At atmospheric pressure, the usual temperature range within which complete liquefaction occurs is between  $-165^{\circ}\text{C}$  and  $-155^{\circ}\text{C}$ . However, since the critical temperature of natural gas is about  $-80^{\circ}\text{C}$  to  $-90^{\circ}\text{C}$ , the gas cannot be liquefied purely by compressing it. It is therefore necessary to use refrigeration processes.

25           Natural gas liquefaction plants are either designed as what are known as LNG baseload plants, i.e. plants for the liquefaction of natural gas for the supply of natural gas as primary energy, or as what are known as peak-shaving plants, i.e. plants for the liquefaction of natural gas to cover peak demand.

30           It is known to cool natural gas by using heat exchangers in which a refrigerant or coolant is used. One well-known method comprises a number of coolant or refrigeration cycles in the form of a cascade.

35           LNG baseload plants are operated as a rule with coolant circuits consisting of a mixture of

hydrocarbons. These mixed refrigerant circuits are more efficient in terms of energy than expander circuits and make it possible, with the high liquefaction capacities of the baseload plants, for correspondingly relatively low energy consumptions to be achieved.

Conventional liquefaction processes using only two refrigerant cycles are limited to about 5 million tons per annum (mtpa) LNG, unless parallel strings within a single train are considered.

The Mixed Fluid Cascade process is known, e.g. from the US patent 6,253,574, and uses three independent refrigeration cycles, which shifts the limit of a real single train concept with proven compressor drivers to above 8 mtpa LNG.

This method is also known from the German published application 197 16 415. With the citing of the German published application 197 16 415, its disclosure content is adopted into the disclosure content of the present Patent Application.

With liquefaction methods of this type, in principle the first coolant circuit serves to provide pre-cooling, the second coolant circuit serves to provide the liquefaction, and the third coolant circuit serves to provide the sub-cooling for the hydrocarbon-rich flow or natural gas respectively.

Between the pre-cooling and the liquefaction, if necessary, the separation of hydrocarbons with higher boiling points takes place. These are at least those components of the hydrocarbon-rich flow or natural gas which would freeze out during the following cooling stage, i.e.  $C_{5+}$  hydrocarbons and aromates. Often, in addition, those hydrocarbons, meaning in this situation in particular propane and butane, which would undesirably increase the calorific value of the liquefied natural gas are also separated out before the liquefaction stage.

This separation of hydrocarbons with higher boiling points usually takes place by provision being made for what is known as an HHC (Heat High Carbon) column, which serves to separate the heavy hydrocarbons as well as benzene out of the hydrocarbon-rich flow which is to be liquefied. A process stage of this kind is likewise described in the German published application 197 16 415 already mentioned; see, for example, Figure 2 and the drawing description relating thereto.

Due to the provision of this separation between those components of the hydrocarbon-rich flow which is to be liquefied, which in the final analysis represents the liquefied product, which is essentially methane and ethane, and those components which are (must be) separated for the reasons given heretofore, at a given pressure of the raw gas the temperature level of the separation of these components from the hydrocarbon-rich flow which is to be liquefied, designated hereinafter as  $C_{3+}$  separation, is set within comparatively narrow limits.

If the first coolant circuit is now used exclusively for the pre-cooling of the hydrocarbon-rich flow which is to be liquefied before this  $C_{3+}$  separation, then a part of the overall compression effect of some 40 to 50 % will necessarily be lost on this, while the remaining compression effect of 60 to 50 % will be divided over the second and third coolant circuits.

In the sense of an economical exploitation of the available compressor and drive units, however, the inventors have realised that it is desirable for the (circuit) compressors of the three circuits to retain approximately the same drive capacity, i.e. in each case about a third of the overall drive capacity. This applies in particular to large liquefaction plants with a liquefaction capacity greater than 5

mtpa, because the number of available compressors and drive units for such orders of magnitude is severely restricted. By standardizing the drive units and compressors of the three coolant circuits, it is possible to maximize the attainable liquefaction capacity of the liquefaction process using tried-and-trusted drive units and compressors respectively.

Thus according to one aspect of the invention there is provided an LNG liquefaction process having first and second refrigeration circuits wherein the second refrigeration circuit is used at least partially for pre-cooling the hydrocarbon-rich stream to be liquefied. Thus the installed power of the gas turbines and starters - at least during normal liquefying operation - can be exploited to the full.

Thus that part of the refrigerant of the liquefaction cycle (LC) may be vaporized under elevated pressure in the precooling section of the process and fed to the LC compressor as a side stream. In this way a substantial load balancing between all three refrigeration cycles can be achieved.

Therefore, according to one aspect of the present invention there is provided a method for the liquefaction of a hydrocarbon-rich flow, in particular of a natural gas flow, whereby the liquefaction of the hydrocarbon-rich flow is effected against a mixed refrigerant circuit cascade consisting of three refrigeration circuits, whereby the first of the three refrigeration circuits serves to provide preliminary cooling, the second refrigeration circuit serves to provide the actual liquefaction, and the third refrigeration circuit serves the sub-cooling of the liquefied hydrocarbon-rich flow, and whereby each refrigeration circuit comprises at least one one-stage or multi-stage compressor, characterised in that at least one part flow of the refrigerant of the second

refrigeration circuit is used for the preliminary cooling of the hydrocarbon-rich flow.

5 From another aspect, the invention provides a method of liquefying a hydrocarbon-rich gas, wherein the gas flows through a cascade of three refrigeration stages, each stage comprising a refrigerant circuit and a compressor, wherein at least part of the flow of refrigerant from the second circuit is used for the preliminary cooling of the hydrocarbon rich gas in the first refrigeration stage.

10 Preferably the part flow of the refrigerant of the second refrigeration (or cooling) circuit, used for the pre-cooling of the hydrocarbon-rich flow is evaporated at a pressure which is higher than the evaporation pressure of the remaining part flow of the refrigerant of the second cooling circuit, and is conducted to the compressor of the second cooling circuit at an intermediate pressure level.

15 Preferably the separation of unwanted components and/or components of the hydrocarbon-rich flow which freeze out during the liquefaction of the hydrocarbon-rich flow takes place before the actual liquefaction of the hydrocarbon-rich flow.

20 Preferably the volumes and/or evaporation pressures of the two part flows of the second cooling circuit are changeable.

25 Preferably at least one part flow of one of the two part flows of the second cooling circuit is used for the provision of cold in the separation unit.

30 The invention therefore provides a load balanced liquefaction process for LNG in which each compressor may have a substantially equal share of the total load, and preferably an equal share. This concept can be applied more widely and hence from another aspect the present invention provides a method of liquefaction comprising a plurality of cooling circuits arranged in a cascade formation, each

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circuit comprising a compressor, wherein each compressor has a substantially equal share of the total load.

5 It will be appreciated that the use of hydrocarbons as refrigerants poses a safety issue and this is particularly significant in the offshore environment, where it is highly undesirable to have large liquid hydrocarbon inventories in what is inevitably a confined space.

10 Floating LNG production, storage and offloading facilities (LNG FPSOs) are now considered a realistic option for remote offshore gas fields that cannot be economically exploited with conventional onshore technology. A floating concept may soon become the  
15 preferred solution for draining deep-water gas reserves.

Therefore the need to increase the safety of such a system is of great importance.

20 One possibility is to use a nitrogen based process, but this has the significant disadvantage that the thermal efficiency is much lower than a hydrocarbon based system. In addition, because nitrogen has a low heat transfer coefficient, a large heat transfer area is required to dissipate the waste  
25 heat from the process into a cooling medium. Consequently, despite the safety hazards involved, hydrocarbon-based refrigeration cycles continue to be used.

30 Another non-flammable and inert refrigerant option is carbon dioxide, which may operate in a vapour compression cycle giving reasonable efficiency. Carbon dioxide has a freezing point of -56.6°C, which restricts the minimum possible  
35 evaporating temperature due to the risk of dry ice formation. Therefore carbon dioxide is an option for the precooling process only. Since most of the hydrocarbon refrigerant inventory is in the

precooling cycle, a change over to CO<sub>2</sub> may still improve the safety of the liquefaction process significantly.

5        Apart from being non-flammable and its high triple point, carbon dioxide is also distinguished from the common hydrocarbon refrigerants for natural gas precooling by its rather low critical temperature (31.1°C), which is comparable to that of ethane (32.3°C).

10        WO 01/69149 discloses the possibility of providing a carbon dioxide precooling circuit in a cascade arrangement with a main cooling circuit.

15        The low critical temperature of CO<sub>2</sub> is a disadvantage since the throttling loss and heat rejection loss in the refrigerating cycle will be larger than for C<sub>3</sub> and C<sub>3</sub>/C<sub>2</sub> mixtures. In addition, the heat transfer loss will be larger than with mixed refrigerant due to constant-temperature evaporation.

20        It has been found that replacing a traditional C<sub>3</sub>/C<sub>2</sub> precooling process, for example that disclosed in US 6,253,574, with an equivalent CO<sub>2</sub> process increases the total power consumption for liquefaction by about 10%, which is considered unacceptable. This consumption increase is due to  
25        the reduction in efficiency of the cycle due to the low critical temperature of carbon dioxide. In addition, the evaporating temperature in the first stage of the CO<sub>2</sub> precooling cycle is only a few degrees higher than the CO<sub>2</sub> triple point. This leads  
30        to operational problems and a danger of dry ice formation.

35        There therefore exists a need for an efficient liquefaction process containing a CO<sub>2</sub> precooling circuit.

      The applicants of the present invention have realised that a carbon dioxide pre-cooling circuit can be combined with the load balanced liquefaction

process described above in order to overcome the above discussed problems with using carbon dioxide.

Therefore, in a preferred embodiment of the present invention the first refrigeration circuit comprises carbon dioxide.

This concept is considered inventive in its own right and therefore, according to another aspect of the present invention there is provided a substantially load balanced mixed refrigerant cascade process comprising a carbon dioxide pre-cooling circuit.

As the liquefaction compressor takes over some of the pre-cooling cycle load, the carbon dioxide circuit can be operated to provide a higher minimum evaporation temperature and thus the risk of dry ice formation is reduced. In addition, as the load of the carbon dioxide cycle is reduced the impact of the lower thermodynamic efficiency of  $\text{CO}_2$  compared with  $\text{C}_2/\text{C}_3$  is alleviated. In a load balanced process where each compressor contributes a third of the total power consumption, the increase in power consumption caused by using  $\text{CO}_2$  can be reduced to only a few percent greater than when using hydrocarbons.

In order to achieve maximum efficiency from the carbon dioxide circuit it is preferable that the carbon dioxide is cooled after condensation to a temperature of  $20^\circ\text{C}$  or less, more preferably to  $15^\circ\text{C}$  or less. This can be achieved using air cooling although preferably cold cooling water is used. As the invention is particularly suited for offshore application the water is preferably sea water, preferably extracted from a deep suitable to give the required low temperature.

Preferably therefore the carbon dioxide pre-cooling cycle includes a sub-cooling heat exchanger installed after the condenser.

Using this method the reduction in total power consumption is great enough to make using a CO<sub>2</sub> pre-cooling circuit a viable option in both on and offshore LNG facilities.

5            Preferably the carbon dioxide cooling circuit comprises three pressure levels in order to improve the thermodynamic efficiency of the process.

10           In order to reduce the internal heat load of the precooling circuit it is preferable that the carbon dioxide is not subcooled in the pre-cooling circuit. This is unlike the second and third cooling cycle refrigerants, the sub-cooling of which increases the efficiency of the process.

15           The higher operating pressure required when using CO<sub>2</sub> means that it may be preferable to use a high pressure casing with the carbon dioxide compressor. More preferably the compressor can be split into two casings and a barrel type casing used for the high pressure stage.

20           According to another aspect of the present invention there is provided a LNG liquefaction process comprising three cascade cycles each driven by a compressor, wherein the compressors are substantially equally loaded and one of the cascade  
25           cycles is a carbon dioxide cycle.

            According to a further aspect of the present invention there is provided a carbon dioxide pre-cooling circuit for LNG liquefaction wherein the carbon dioxide has a minimum evaporation temperature  
30           of no less than -50°C, preferably no less than -40°C and most preferably no less than -35°C.

            Preferred embodiments of the present invention shall now be described, by way of example only, with reference to the following drawings, in which:

35           FIG 1 shows a load balanced liquefaction process in accordance with a preferred embodiment of the invention;

FIG 2 show an alternative embodiment of a load balanced process;

FIG 3 shows a graph of overall power demand as a function of a reference temperature;

5 FIG 4 shows a load balanced liquefaction process containing a carbon dioxide pre-cooling circuit;

FIG 5 shows hot/cold composite curves for the processes shown in FIGs 2 and 4; and

10 FIG 6 shows a comparison of refrigerant inventories of the processes shown in FIGs 2 and 4.

In FIG 1 the cooling and liquefaction of the hydrocarbon-rich flow, which is conducted via line 1 to the heat exchanger E1, are effected against a cooling agent mixture (or mixed refrigerant) circuit cascade, consisting of three cooling agent mixture circuits. These as a rule have different compositions, such as are described, for example, in the aforementioned German published application 197 16 415.

20 The hydrocarbon-rich flow which is to be liquefied is cooled in the heat exchanger E1 against the two evaporating cooling agent mixture flows 4b and 4d of the first mixture circuit 4a to 4e, and then conducted via line 1a to a separation unit S, represented simply as a box.

25 In this separation unit S the  $C_3$  separation described heretofore takes place, whereby the components separated out of the hydrocarbon-rich flow which is to be liquefied are drawn off from the separation unit S via line 1b.

30 According to one advantageous embodiment of the method according to the invention, not shown in the drawing, at least one part flow of one of the two part flows 3b and 3d of the second cooling agent mixture circuit 3a to 3e, which will be discussed in greater detail hereinafter, is used for the provision of

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cooling in the separation unit S. In this situation, the choice of which of the two part flows 3b and/or 3d at least one part flow is drawn from in turn for this provision of cooling is determined by the temperature level(s) required in the separation unit S.

The hydrocarbon-rich flow to be liquefied is then conducted via line 1c to a second heat exchanger E2, and is liquefied in this against the evaporating cooling agent mixture flow 3b of the second cooling circuit 3a to 3b.

Once liquefaction has taken place, the hydrocarbon-rich flow is conducted via line 1d to a third heat exchanger E3, and is subcooled here against the cooling agent mixture flow 2b of the third cooling circuit 2a to 2c. The subcooled liquid product is then conducted via line 1e to its further use.

As can be seen from the drawing, each of the three cooling circuits 2a to 2c, 3a to 3e, and 4a to 4e, has a compressor, V2, V3, and V4 respectively. Not shown in the drawing are the corresponding drives for these compressors V2, V3, and V4. In addition, the coolers or heat exchangers which are located downstream of the compressors V2, V3, and V4 respectively are not shown in the drawing, in which the coolant mixture is cooled against a cooling medium, such as water.

The refrigerant mixture of the first mixture circuit, compressed in the compressor V4, is conducted via line 4a to the heat exchanger E1, and is divided here into two part flows 4b and 4d after cooling has taken place. The refrigerant mixture in these part flows 4b and 4d, after throttling has been effected in the valves d and e or expansion devices, is evaporated to different pressure levels in the heat exchanger E1 and then conducted via line 4c or 4e to the compressor V4 before the first stage (part flow 4c) or to an intermediate pressure level (part flow 4e).

The refrigerant mixture of the second cooling circuit 3a to 3e, compressed in the compressor V3, is conducted via line 3a through heat exchangers E1 and E2, and is cooled in these. That part flow 3b of this refrigerant mixture flow, which is conducted through heat exchanger E2, after expansion in valve b, is evaporated in heat exchanger E2 against cooling process flows, and is then conducted via line 3c to the intake stage of compressor V3.

According to the invention, a part flow 3d of the refrigerant mixture of the second refrigerant mixture circuit 3a to 3e is drawn off after the heat exchanger E1, expanded in valve c, and then evaporated in heat exchanger E1 against cooling process flows, before being conducted via line 3e, at an intermediate pressure level, to the circuit compressor V3. Accordingly, the refrigerant mixture part flow 3d referred to, according to the invention, makes a contribution to the pre-cooling of the hydrocarbon-rich flow in heat exchanger E1.

In order for this to be achieved, the part flow 3d of the refrigerant mixture of the second refrigerant mixture circuit 3a to 3e, used for the pre-cooling of the hydrocarbon-rich flow, must be evaporated at a pressure which is higher than the evaporation pressure of the refrigerant mixture part flow 3b of the second cooling agent mixture circuit 3a to 3e.

By selecting the intermediate pressure at which the refrigerant mixture part flow 3e is evaporated and conducted to the compressor V3, and by regulating the volume distribution of the two refrigerant mixture part flows 3b and 3d, the distribution of the cooling capacity of the second mixture circuit onto the heat exchangers E1 and E2, and therefore to the pre-cooling and liquefaction of the hydrocarbon-rich flow which is to be liquefied, can be adjusted almost at will.

If, for example, 40 % of the total drive capacity is required for the pre-cooling and 60 % for the liquefaction and subcooling of the hydrocarbon-rich flow, then, with the concept and method according to the invention, one compressor is used in each case with a third of the total drive capacity in the first and third refrigerant mixture circuit, i.e. for the pre-cooling as well as for the subcooling of the hydrocarbon-rich flow which is to be liquefied. The compressor of the second refrigerant mixture circuit is operated according to the invention in such a way that it uses 20 % of its capacity, and consequently 6.66 % of the total capacity, for pre-cooling, while the remaining 80 %, i.e. 26.66 % of the total capacity, is used for liquefaction.

The method according to the invention for the liquefaction of a hydrocarbon-rich flow, in particular of a natural gas flow, accordingly makes possible the economical exploitation of the available compressors and drive units, because the (circuit) compressors of the three mixture circuits obtain approximately the same drive capacity, i.e. a third of total capacity in each case. Accordingly, large liquefaction plants in particular, with a liquefaction capacity greater than 5 million tonnes LNG per year, can be operated substantially more economically, since, by standardizing the drives and compressors of the three cooling circuits, the achievable liquefaction capacity of the liquefaction process can be maximised with the use of tried-and-trusted drive units and compressors.

FIG 2 shows an alternative version of the load balanced process. As with FIG 1 the pre-cooling cycle C10 comprises a first circuit driven by a first compressor V10 and one part 22 of the refrigerant stream 21 from the second cycle C20. Three General Electric MS 7121 EA (Frame 7) are used to drive the compressors V10, V20, V30. If highest availability is

of the essence, the three refrigeration cycles can be designed with two times 50% gas turbine/compressor trains. In this case six GE MS 6581 B (Frame 6) would replace the three Frame 7s.

5 All LNG plants require the extraction of at least of those hydrocarbons, which would freeze in the LNG under storage conditions (e.g. aromatics and C<sub>5</sub>plus). In an LNG plant usually precooling is considered as first cooling step between ambient  
10 temperature and extraction of the mentioned hydrocarbons.

It should be emphasised that the method according to the invention can be combined with all known separation methods considered to be prior art  
15 for relatively high-boiling hydrocarbons.

The precooling portion of the overall power demand of all refrigeration compressors for the two gases defined in Table 1 is shown in FIG 3 as a function of a reference temperature. This is the  
20 temperature, under which all main process streams (natural gas, refrigerant fluids) enter into the cryogenic heat exchangers.

Table 1

25

mol%	Lean	Rich
	Gas	Gas
N <sub>2</sub>	5.00	5.00
CH <sub>4</sub>	88.93	84.07
C <sub>2</sub> H <sub>6</sub>	3.96	5.58
30 C <sub>3</sub> H <sub>8</sub>	1.37	2.73
C <sub>4</sub> H <sub>10</sub>	0.48	1.34
C <sub>5</sub> H <sub>12</sub>	0.17	0.65
C <sub>6</sub> H <sub>14</sub>	0.06	0.32
C <sub>7</sub> H <sub>16</sub>	0.02	0.16
C <sub>8</sub> H <sub>18</sub>	0.01	0.08
35 Benzene	0.01	0.08
	100.0	100.0
	0	0

5           The lower the reference temperature and the  
richer the gas the smaller the required compressor  
power for precooling becomes. This situation can be  
addressed reasonably well by designers of dual flow  
liquefaction processes, if the power mismatch between  
precooling and liquefaction plus subcooling is  
compensated by helpers for the gas turbines.

10           A process with three refrigeration cycles such  
as the MFC<sup>o</sup> process offers a much wider field for even  
load distribution between the cycles. If part of the  
refrigerant of the liquefaction cycle C20 is  
vaporized under elevated pressure in the precooling  
section C10 and is fed to the LC compressor V20 as  
15 side stream 22, a perfect load balancing between all  
three refrigeration cycles can be achieved. This  
feature is a major aspect of a cost effective design  
for large production capacities. As all three (3)  
cycles are symmetrically driven this arrangement is  
referred to as MFC<sup>o</sup>s3.  
20

          In order to provide actual figures for a  
realistic process design a large LNG train has been  
studied. On the basis of the lean gas composition  
with 62 bar and 35 deg C at the inlet to precooling a  
25 conceptual process design was made. The refrigeration  
compressors are driven by Frame 7's with additional  
20 MW on each shaft, which have been recruited from  
the starter/helpers. The resulting LNG rundown  
amounts to 8.5 mtpa at 333 stream days, which is  
30 accompanied by an additional quantity of 0.4 mtpa NGL  
(C<sub>3</sub> plus hydrocarbons). The specific energy consumption  
of the refrigeration compressors is 259 kWh/t<sub>LNG</sub>.

          Unlike the embodiment of FIG 1, the final  
compressor V30 of FIG 2 is split into two casings  
35 V31, V32. The second casing V32 is designed to deal  
with high pressures at which the multistage  
compressor operates.

In FIG 4 the precooling circuit C10 of FIG 2 has been replaced with a pre-cooling circuit C100 which comprises a carbon dioxide stream 101. After compression and condensation/subcooling the stream 101 is split into three separate streams, 102, 103, 104 which are then expanded to different pressures. This compensates for the constant temperature evaporation of CO<sub>2</sub>. Unlike hydrocarbon streams 201, 301 the carbon dioxide stream 101 is not sub-cooled by the pre-cooling heat exchanger E100 prior to expansion in order to reduce the internal heat load of this exchanger.

Owing to the higher operating pressure, the CO<sub>2</sub> precooling compressor V100 is split into two casings, V110, V120 with a barrel type casing V120 for the high-pressure stage. Not shown in the figure is an additional subcooling heat exchanger on the CO<sub>2</sub> circuit, using seawater to subcool the liquid refrigerant after the condenser, in order to improve process efficiency.

As with the previous embodiments "load balancing" is achieved by allowing the liquefaction compressor V200 to take over some of the precooling cycle load, leading to a "symmetrical" process.

Process simulations of the above embodiment as shown in FIG 4 and FIG 2 gave power requirement data as shown in Table 2, using the design data as shown in Table 3. As a result of the load-balanced process, the power input to the CO<sub>2</sub>-precooled case was only 4.4% higher than the baseline. For a given maximum available power as defined by the hydrocarbon process case, this would correspond to a LNG capacity of 95.6% with CO<sub>2</sub> precooling unless more driver capacity is installed.

Table 2

	CO <sub>2</sub> precooling	C <sub>2</sub> /C <sub>3</sub> precooling	
Total shaft power	162.7 (104%)	155.8 (100%)	MW
Precooling compressor	49.6	47.6	MW
Liquefaction compressor	50.5	47.7	MW
Subcooling compressor	50.5	48.5	MW
Other power consumers	12.1	12.0	MW

Table 3

LNG production capacity	5.8	mtpa
Gross calorific value of LNG	40	MJ/Sm <sup>3</sup>
Feed gas pressure (liquefaction inlet)	69	bar
Sea cooling water temperature	5	°C

5                    Temperature profiles in the form of hot/cold  
 composite curves for the two cases are shown in  
 Figure 5. The three CO<sub>2</sub> precooling temperature levels  
 are easily observed in the left diagram. The highest  
 pressure level to the liquefaction compressor is also  
 10           considered part of precooling. Changes in the  
 subcooling process are minimal between the two cases.

Layout, size and weight of the liquefaction  
 module with CO<sub>2</sub> precooling were compared to the  
 baseline hydrocarbon case (that shown in FIGs 1 and  
 15           2). Among the factors that contributed to reduce the  
 equipment footprints and give smaller dimensions with  
 CO<sub>2</sub> were reduced precooling compressor suction drum  
 sizes and smaller precooling piping dimensions.  
 Additional equipment caused by the third precooling  
 20           pressure level/drum and the installation of a  
 refrigerant subcooler made the net reduction in  
 footprint area marginal, however. The plate-fin heat  
 exchangers were reduced in size due to larger LMTD  
 (Logarithmic Mean Temperature Difference) and less  
 25           internal duty. Some of the major pipe sizes in the

liquefaction and subcooling circuit did not change much, and it is these pipes that to a large degree set out the deck heights, so no changes were envisaged relative to deck elevations. In total, it was concluded that the liquefaction module size would be the same for both cases, although a few square meter reduction is possible for the CO<sub>2</sub> case. The weight of the module dropped by 100 tons.

A major safety concern of the LNG process with hydrocarbon precooling, especially when applied offshore, is the possible formation of a flammable and explosive hydrocarbon/air mixture in case of a major leakage in one of the refrigerant circuits. Thus, the minimization of hydrocarbon refrigerant inventory is very important in terms of safety.

As may be observed from the FIG 6, the HC refrigerant inventory is reduced by about 70% in the CO<sub>2</sub>-precooled process. The reduced hydrocarbon charge is positive in relation to loss prevention and to the availability of the three main safety functions of the LNG barge, which are (i) main structural strength, (ii) main escape routes, and (iii) means of evacuation.

If the molecular weight of the hydrocarbon refrigerant is higher than that of air, a flammable cloud can accumulate inside or between the modules, and on the deck surfaces. Thus, in addition to minimizing the total hydrocarbon inventory it is of special importance to eliminate the heavier components, especially propane (52% heavier than air), but also ethane (4% heavier than air). By replacing the hydrocarbon precooling with CO<sub>2</sub>, all propane is eliminated from the liquefaction module, and even though ethane is present in the liquefaction and subcooling refrigerants, both these mixtures have a molar mass that is lower than air.

From the above results it has been found that the introduction of CO<sub>2</sub> precooling in a load-balanced MFC<sup>s</sup>3 process does not give a significant increase in specific power requirement, or equipment size/weight/cost, while the safety of the process can be improved. It should be kept in mind, however, that the above results are based on cold seawater cooling. Cold water is of course a benefit for all LNG processes, but due to the low critical temperature of CO<sub>2</sub>, this process is especially sensitive to increased cooling water inlet temperature. The CO<sub>2</sub> process efficiency would drop more than a corresponding HC process if the coolant (sea water or air) temperature were increased.

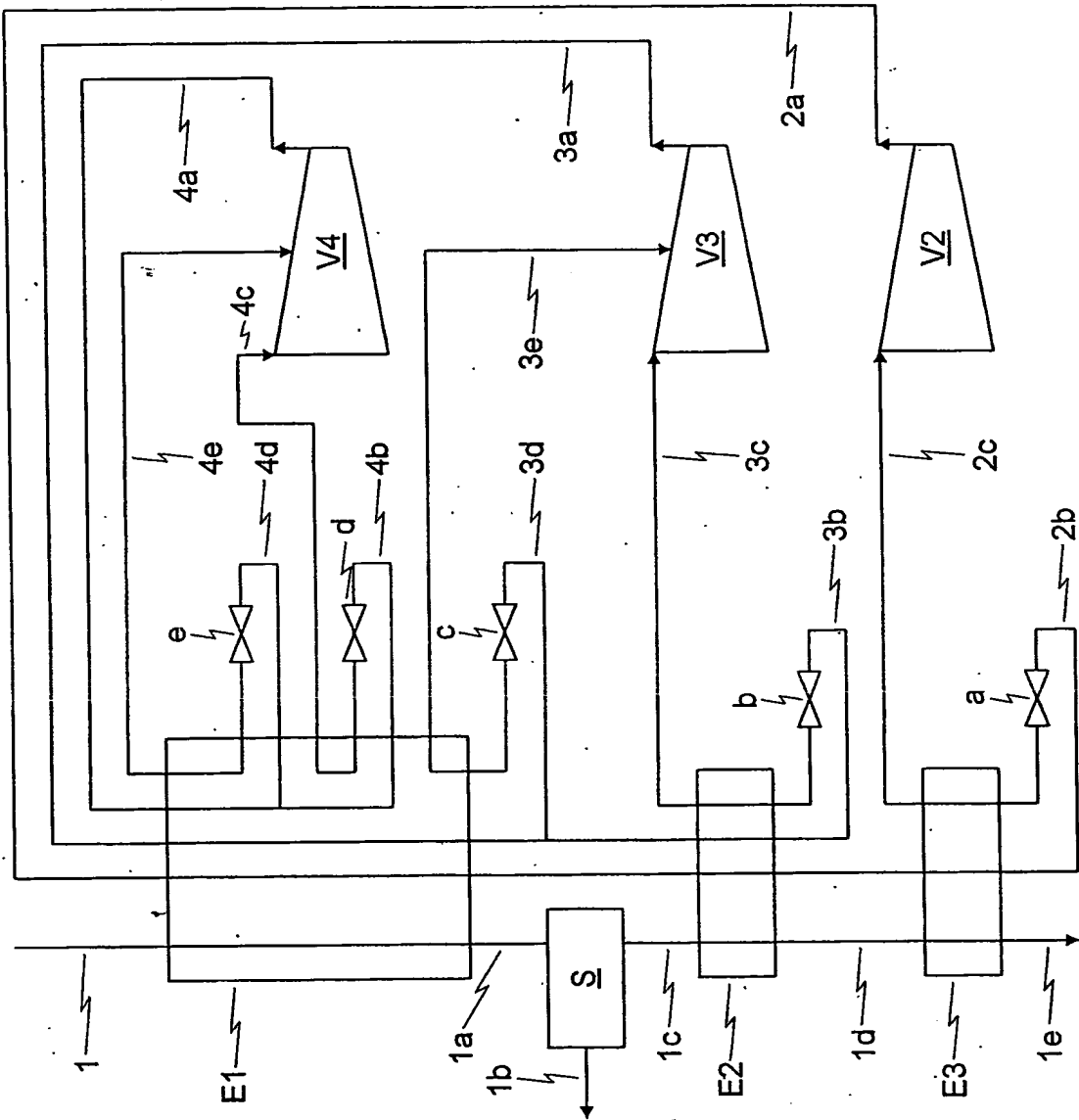


Fig 1

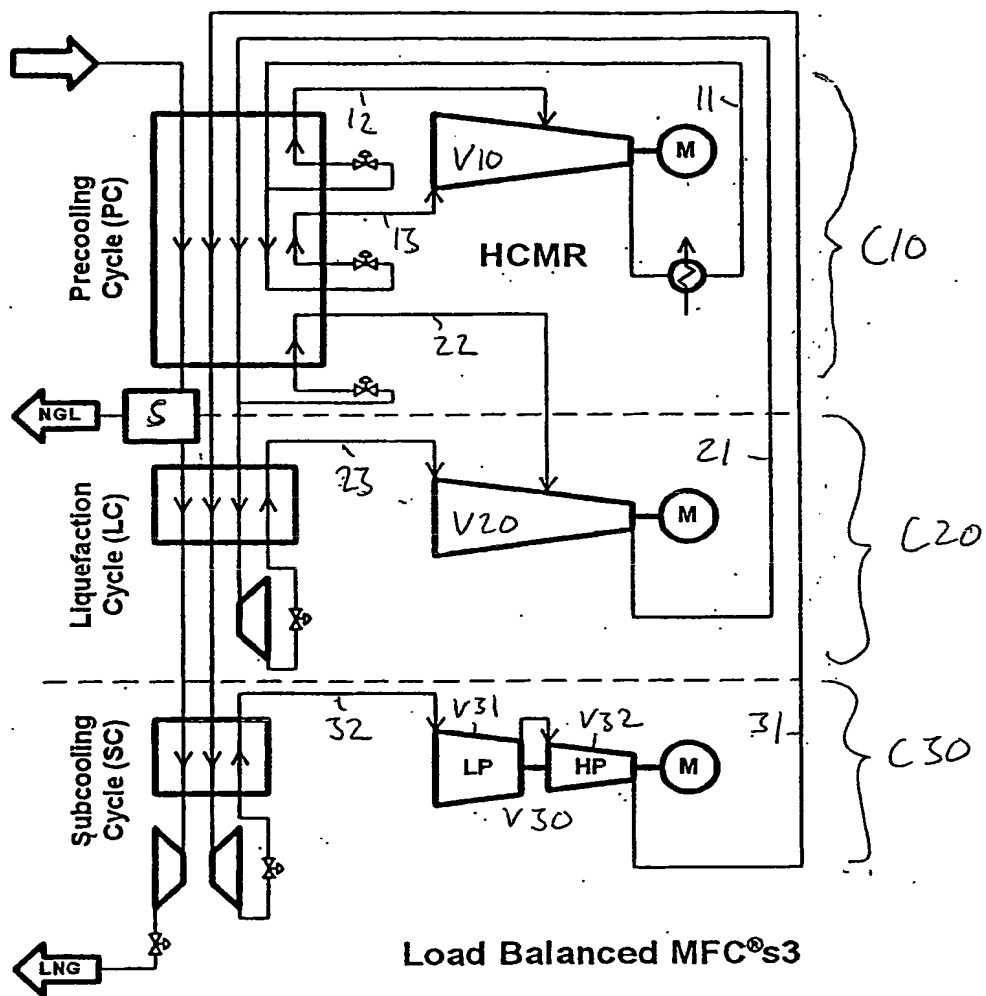


FIG 2

315

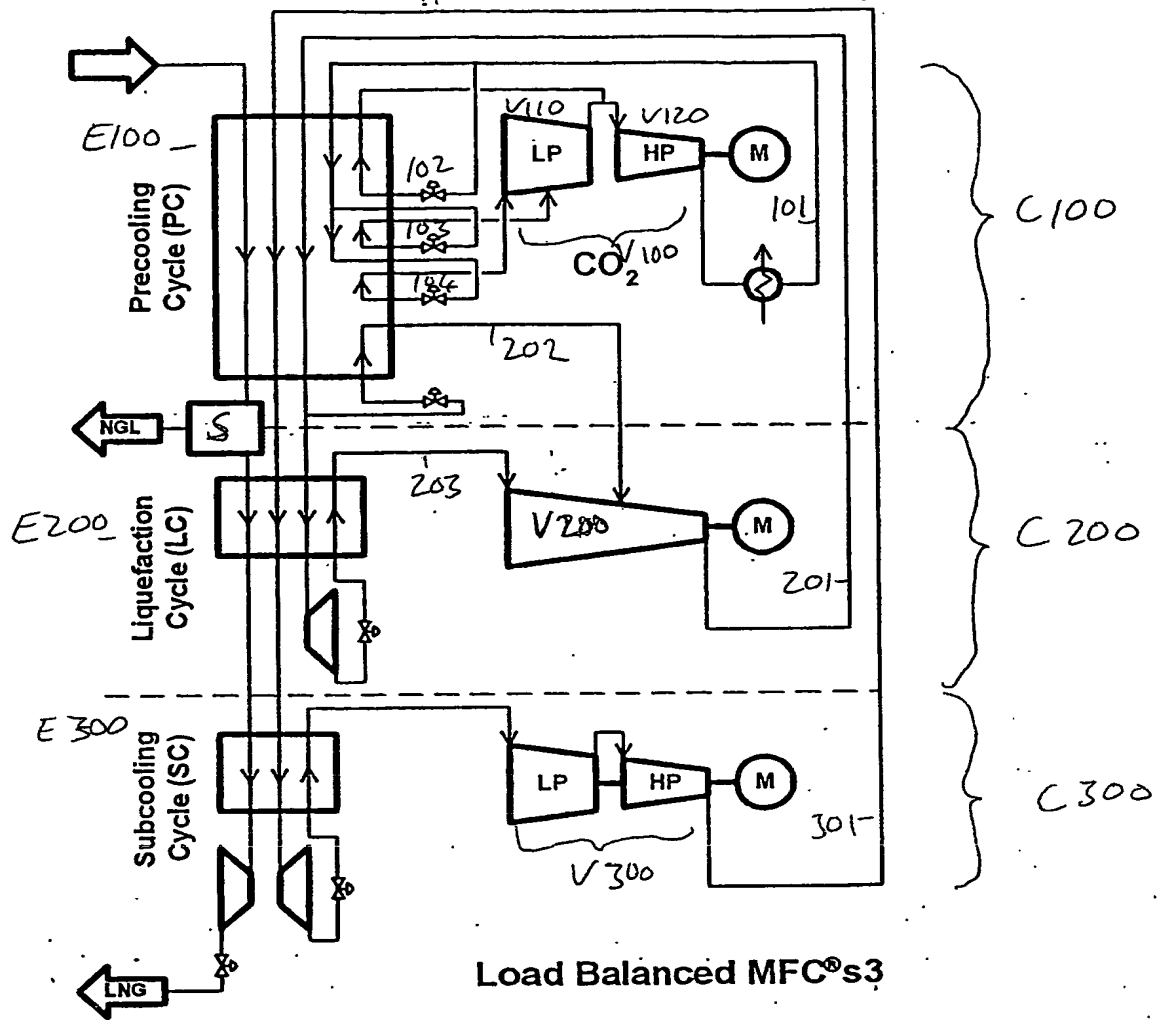


Fig 4

415

### Precooling Power Portion

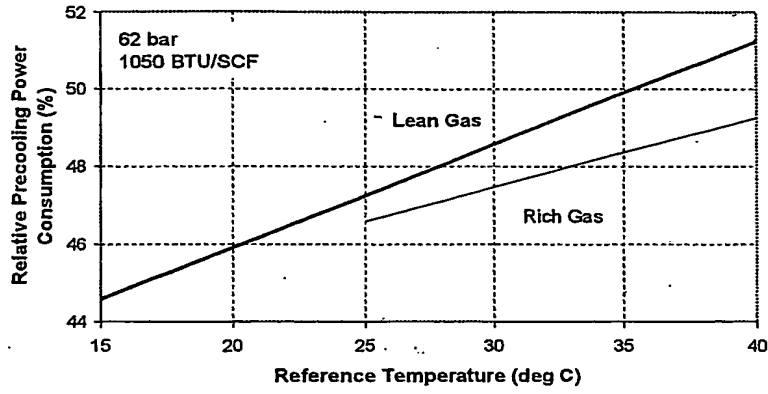


FIG 3

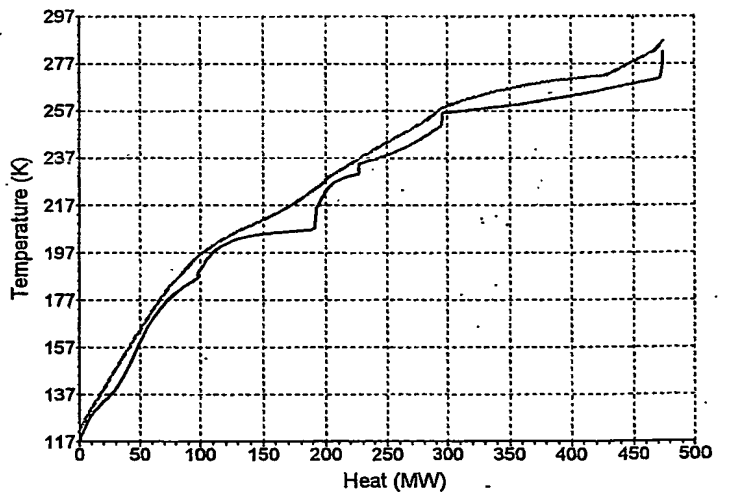
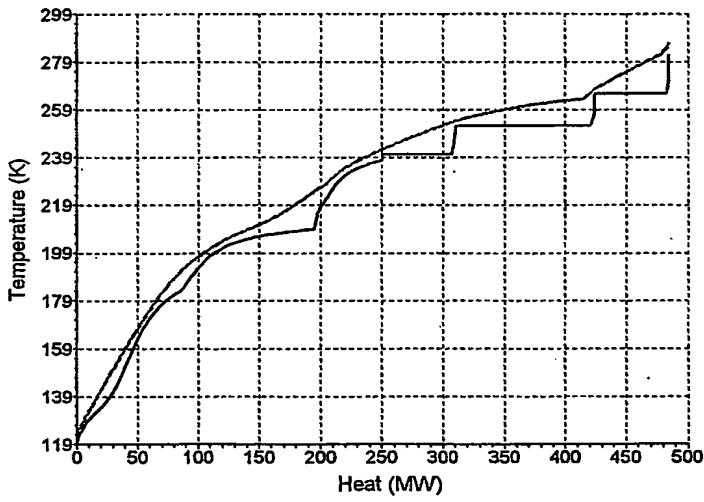


FIG 5

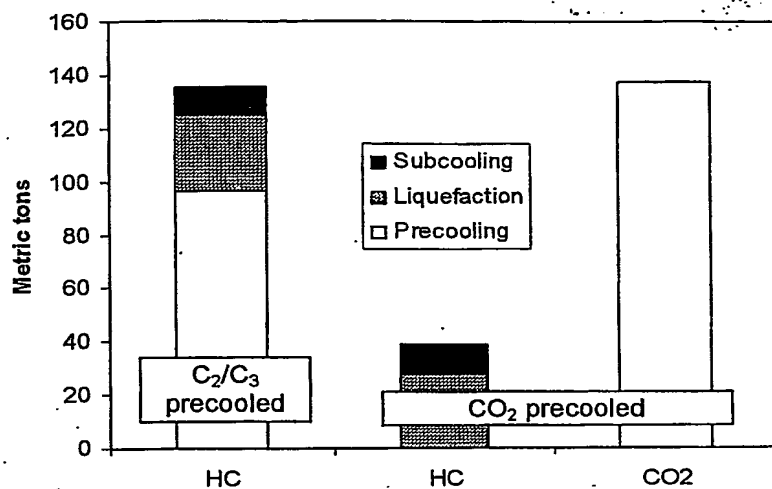


FIG 6

PCT/GB2004/004047



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